PB 189 377

FINAL REPORT

ON

SULFUR DIOXIDE SCRUBBERS

STONE & WEBSTER - IONICS PROCESS

CONTRACT NO. CPA 22-69-80

FOR

DIVISION OF PROCESS CONTROL ENGINEERING

NATIONAL AIR POLLUTION CONTROL ADMINISTRATION

U. S. DEPARTMENT OF HEALTH, EDUCATION AND WELFARE

JANUARY 1970

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STONE & WEBSTER ENGINEERING CORPORATION

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225 FRANKLIN STREET, BOSTON, MASSACHUSETTS 02107

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January 30, 1970

DESIGN
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REPORTS
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CONSULTING
ENGINEERING

Mr. Jonathan P. Earhart
New Process Development Unit
Process Control Engineering Division
National Air Pollution Control Administration
5710 Wooster Pike
Cincinnati, Ohio 45227

Dear Mr. Earhart:

In accordance with the authorization contained in Contract No. 22-69-80, dated May 1, 1969, we have made a study to determine the most economical scrubber for use in the removal of sulfur dioxide from flue gas produced by a coal burning power plant.

The scrubber would be specifically designed to operate as part of a system using the Stone & Webster/Ionics sulfur dioxide recovery process as described in the proprietary Stone & Webster proposal dated July 22, 1968, previously presented to NAPCA.

The design described in the S&W proposal was based on removing sulfur dioxide from the flue gases produced in a coal-fired power station rated 1,200 Mw and consisting of four boilers. The flue gas from one boiler amounts to 880,000 actual cfm at 300 F and contains 0.3 mole percent sulfur dioxide. Fly ash loading is 0.05 grains per cu ft. The electrolytic regeneration system will produce a feed stream to the scrubber system which contains about 4.4 moles of NaOH plus 2.7 moles of Na_2SO_4 per 100 moles of water.

The equipment provided consisted of a tower with water sprays for cooling the flue gas and removing some residual fly ash, a sulfur dioxide removal section comprised of two packed sections of Johnstone wood slat packing, an induced draft blower, scrubbing liquor circulation pumps, and necessary connecting duct work, piping and instruments.

To determine whether any other type of scrubber is superior to the Johnstone packed tower, quotations were requested from 23 manufacturers of various types of gas scrubbers. Of the quotations received, one in each of the following classes of scrubbers was selected for estimation:

Stone & Webster Ripple Tray tower
Floating ball contactor
Intalox Saddle packed tower
Impingement grid packed tower
Venturi contactor
Cyclonic contactor

An attempt was made to determine the percentage oxidation of sulfites to sulfate for each type of scrubber. Values from the literature and quoted values from vendors were used. No positive conclusions regarding degree of oxidation for each type could be reached.

Investment and operating costs were estimated for each type of scrubber system studied. The conclusions reached can be summarized as follows:

1. Large single scrubber installations are least expensive to install and overall least costly to operate.

2. Proprietory equipment such as venturi, cyclonic and floating ball scrubbers is impractical in the S&W/Ionics process at the present time. Equipment manufacturers will have to produce units with much larger throughput to be competitive with packed towers.

3. The differences in large single scrubbers such as packed and Ripple Tray towers are not great, although the Johnstone wood grid packed towers showed some advantages over the other types.

For further details, we refer you to the following pages of this report.

Yours very truly,

E. G. Lowrance

E. G. Lowrance
Project Director

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INTRODUCTION

SCOPE OF WORK

The National Air Pollution Control Administration (NAPCA) of the U. S. Department of Health, Education and Welfare (HEW) engaged Stone & Webster Engineering Corporation to make a study to determine the most economical scrubber for the removal of sulfur dioxide from flue gas produced by a coal burning power plant. The scrubber would be specifically designed to operate as part of a system using the Stone & Webster/Ionics sulfur dioxide recovery process as described in the proprietary Stone & Webster (S&W) Proposal, dated July 22, 1968, which was presented to NAPCA.

The S&W/Ionics sulfur dioxide recovery process includes a sulfur dioxide scrubber in which sulfur dioxide is absorbed in a sodium hydroxide solution to form a sodium sulfite/sodium bisulfite solution which is sent to a neutralization stage.* The economics of the process are such that a high bisulfite-to-sulfite ratio and a minimum oxidation to sulfate are desired. The installed and operating costs included in the S&W Proposal of July 22 1968, were based on using a sulfur dioxide scrubber provided with Johnstone type wood slat packing as described in the paper by Johnstone & Singh of March 1937, I.E.C., Vol. 29.

Since it seemed probable that an improved means of scrubbing might have been developed since Johnstone's work of 1937, this study was authorized to accumulate, organize and report pertinent engineering information on various types of scrubbers which would be technically feasible for use as a chemical contactor in the S&W/Ionics sulfur dioxide removal and recovery process.

The S&W/Ionics Process underwent pilot plant tests during 1967 at the Gannon Station of Tampa Electric Co. in Tampa, Florida, where the operation as a whole and critical components were evaluated, using actual flue gas from a boiler fired by pulverized coal. The pilot unit was sized to process about 200 cfm of flue gas containing approximately 0.25 volume percent sulfur dioxide. The equipment included a flue gas quench which cooled the gas to about 120 F and removed most of the fly ash and which was followed by an absorber. Since our primary purpose was to evaluate the electrolytic cell, only a Venturi type scrubber and later a spray type scrubber were tested. Figure 1 illustrates the flow.

The scope of the study which is described in this report did not allow time or money for any experimental work. It is anticipated that the results of the study will narrow the selection of scrubber types to one or two and that some experimental work may be necessary before the detailed design of a prototype begins.

**The neutralized spent liquor is then regenerated in an electrolytic cell for reclamation of the caustic for reuse in the process.*

DESIGN BASIS

The design described in the S&W proposal was based on removing sulfur dioxide from the flue gases produced in a coal-fired power station rated at 1,200 Mw and consisting of four boilers. The flue gas from each boiler amounts to 880,000 actual cfm at 300 F and contains 0.3 mole percent sulfur dioxide. Fly ash loading is 0.05 grains per cu ft. The electrolytic regeneration system will produce a feed stream to the scrubber system which contains about 4.4 moles of NaOH plus 2.7 moles of Na_2SO_4 per 100 moles of water.

The equipment provided consisted of a tower with water sprays for cooling the flue gas and removing some residual fly ash, a sulfur dioxide removal section comprised of two packed sections of Johnstone wood slat packing, an induced draft blower, scrubbing liquor circulation pumps, and necessary connecting duct work, piping and instruments. Because of the low gas pressure drop in the system, the induced draft fan was placed at the outlet of the sulfur dioxide scrubber.

TYPES OF SCRUBBERS STUDIED

For the present study of sulfur dioxide scrubbers, the same general arrangement of equipment has been retained except that the fan is located upstream of the quench and fly ash removal equipment because of the higher pressure drop associated with most of the other scrubbers studied. This requires handling a larger volume of gas at 300 F, but avoids corrosion problems since the flue gas is above its water dew point.

To determine whether any other type of scrubber is superior to the Johnstone packed tower, quotations were requested from 23 manufacturers of various types of gas scrubbers. Of the quotations received, one in each of the following classes of scrubbers was selected for estimation:

Stone & Webster Ripple Tray tower
Floating ball contactor
Intalox Saddle packed tower
Impingement grid packed tower
Venturi contactor
Cyclonic contactor

In some cases, the quench section and the absorption section use the same type of scrubber, but occasionally a combination was used. In one case, for example, a spray quench followed by a Stone & Webster Ripple Tray tower was used.

An attempt was made to select equipment which would achieve 90 percent sulfur dioxide removal. Since no supporting data were supplied by the vendors, at least two actual contact stages are provided in all cases.

VENDORS CONTACTED AND REPLIES

Table 1 lists the vendors contacted and indicates those who declined to quote for one reason or another. Generally, those who declined to quote stated that the gas volume to be handled was too large for their equipment. Of the 23 companies contacted, nine declined to quote and one (Research-Cottrell, Inc.) quoted on the fly ash removal stage only. Thus, 13 quotations plus the S&W Ripple Tray design were left for consideration.

Since a 90 percent sulfur dioxide removal was a necessary requirement of this evaluation, only those vendors offering a minimum of two contact stages were considered as being acceptable even though no supporting data were supplied by the vendors to substantiate this selection.

Table 2 shows the type of equipment quoted for the quench/fly ash removal section. The type of equipment quoted for the sulfur dioxide absorption section is shown in Table 3.

The UOP Air Correction Division declined to quote because they believe that low liquid velocities and very high retention times are required in producing a liquor high in bisulfite. This is in opposition to the operating principle of their Turbulent Contact Absorber (TCA) scrubber.

Most of the vendors offered to test their equipment or to cooperate in test work preliminary to a design of a prototype unit.

DISCUSSION

PREQUENCH AND FLY ASH REMOVAL

Flue gas leaving the electrostatic precipitators will normally be at about 300 F and about 0 psi gage. Fly ash loading is expected to be about 0.05 grains per cu ft. Expected size distribution of the fly ash is as follows:

Size, Microns	*Wt %*
0-10	45
11-20	23
21-45	20
45 and over	12

An inspection of the proposals received shows that quenching of the flue gas from 300 F to about 130 F can be accomplished. However, only one vendor offered a guarantee on the extent of cooling.

The extent of fly ash removal is less clear. Claims of the vendors varied from 90 to 99 percent removal of the contained fly ash to no estimate at all. The vendor with the highest rate offered removal of 99 percent of the fly ash over 1 micron in size. Supporting data were not submitted by any vendor. We believe some ash will pass through the quench section and will be captured in the sodium sulfite/sodium bisulfite solution. Thus, filtration will be required before the solution can be sent to the electrolytic cells. Therefore, some incremental increase in investment and operating costs will result if fly ash removal in the quench section is lower than expected. However, due to the indefinite nature of the claims, this factor was not included in our estimated costs for this study.

SULFUR DIOXIDE ABSORPTION

Efficiency of Sulfur Dioxide Removal

To meet local air pollution standards that are being proposed, it is necessary to reduce the sulfur dioxide concentration in flue gas to a value equivalent to firing coal with 1 percent sulfur. For a 4 percent sulfur content coal, this would be equivalent to removing 75 percent of the sulfur dioxide produced. Since the S&W/Ionics process recovers sulfur dioxide as a salable product, we have based our designs on a recovery of 90 percent of the sulfur dioxide in the flue gas. Based on a sulfur dioxide concentration at the outlet of the electrostatic precipitators of 0.3 volume percent or 3,000 ppm, this would result in a reduction to 300 ppm by volume. This high recovery not only adds to the salable sulfur dioxide but also permits the utility to meet more rigid standards which may be imposed in the future by local or Federal authorities.

Calculations based on data developed by Johnstone and Singh[22] indicated that about 8 ft of slat packing would be required for this service. Since there is insufficient fresh caustic solution to wet the packing adequately, recirculation of scrubbing liquor is required. Thus, we have divided the packing into two stages of 6 ft each, with trap-out trays and recirculation pumps for each stage. Johnstone and Singh data indicated that, for the wood slat packing configuration used, the height of a transfer unit was approximately 1.6 ft in this service. We have provided 7.4 transfer units as compared with 5.0 units which we have calculated as the required amount.

We do not have a method for comparing the performance of Johnstone wood slat packing and the other types of contactors offered by the vendors solicited. None of the vendors supplied information to substantiate the claims for the performance of his equipment. Most vendors believed that one or, at most, two stages would be adequate. Our own calculations indicate that two packed sections would be required to obtain 90 percent removal, with three packed sections needed for 95 percent removal. One contacting stage would not be adequate because of the high vapor pressure of sulfur dioxide above bisulfite solutions. Most vendors have offered to do test work or to cooperate in test work to determine the number of contacting stages required. Such test work could be carried out at vendors' laboratories or in cooperation with a utility. Discussion of a testing program has been deferred until a prototype unit is under consideration.

Utilization of Caustic

The electrolytic regeneration system will produce a feed to the absorption system which will contain about 2N or 8 wt % sodium hydroxide (pH of about 14). Since all the scrubbers being studied require recirculation of absorption liquor in order to provide adequate liquid loading, this caustic feed will be added to recirculated liquid at the top stage. The amount of free caustic in the combined liquid to the top stage is a function of the amount of recirculation to the stage and the amount of sulfur dioxide absorption in the stage.

The principal reactions that occur in the absorber are as follows:

$$SO_2 + H_2O \rightleftharpoons H_2SO_3$$
$$CO_2 + H_2O \rightleftharpoons H_2CO_3$$
$$2\,NaOH + H_2CO_3 \rightleftharpoons Na_2CO_3 + 2\,H_2O$$
$$Na_2CO_3 + H_2CO_3 \rightleftharpoons 2\,NaHCO_3$$
$$2\,NaOH + H_2SO_3 \rightleftharpoons Na_2SO_3 + 2\,H_2O$$
$$Na_2SO_3 + H_2SO_3 \rightleftharpoons 2\,NaHSO_3$$
$$Na_2CO_3 + H_2SO_3 \rightarrow Na_2SO_3 + CO_2\uparrow + H_2O$$
$$NaOH + NaHSO_3 \rightarrow Na_2SO_3 + H_2O$$

It is expected that there will be little or no free caustic in the combined feed to the top section. This stream will contain mostly Na_2CO_3 and Na_2SO_3. As the liquid flows down through the tower, CO_2 will be displaced by SO_2 and Na_2SO_3 will be partially converted to $NaHSO_3$.

The vapor pressure of sulfur dioxide over neutral sodium sulfite is essentially zero, while over mixtures of sulfite/bisulfite the vapor pressure is a function of the sodium bisulfite concentration. Johnstone[23] found it convenient to characterize the relative amounts of sulfite and bisulfite in a solution by using the ratio of S/C or moles of dissolved SO_2 per mole of available sodium ion (S/C = 0.5 for Na_2SO_3; S/C = 1.0 for $NaHSO_3$). The relationships between the vapor pressure of sulfur dioxide and S/C are discussed in Appendix A.

The absorption system should have sufficient contact stages to produce a liquor which contains a high ratio of S/C and should have adequate contacting in each stage to approach equilibrium between gas and liquid. It is also essential to provide a system in which entrainment from stage to stage is minimized. A ratio of S/C of 0.9 or higher in the scrubber product is required to achieve the most economical cell operation. The pH of this solution leaving the final stage should be about 3.5.

None of the vendors who submitted proposals offered any data that would be useful in determining the extent of conversion to sodium bisulfite. Actually, not much information could be expected from the vendors since caustic or sodium carbonate has not been reported in commercial use to extract sulfur dioxide from flue gas. Recovery of sulfur dioxide from gases produced by combustion of sulfur is practiced in the paper industry as part of a magnesium sulfite pulping process, but magnesium hydroxide, rather than sodium hydroxide, is used as the absorbent. In other sulfite pulping processes, either water or an alkaline solution is used to absorb sulfur dioxide from the gases. In these cases sulfur dioxide is present in much higher concentrations than those appearing in a flue gas from a coal burning power plant.

If vendors could have provided data supporting the claimed sulfur dioxide removal, the ratio of S/C could be calculated. It is our opinion that none of the claimed figures can be used in a system where separate recirculation to each contact stage is used.

Since many of the vendors have test facilities, it should be possible to determine the maximum concentration of bisulfite during the design stage for a prototype unit, but until this is done we will not be able to evaluate this factor.

OXIDATION

The contract requires that an estimate of the percentage of sulfur dioxide transferred from gas to liquid which is oxidized shall be made for each scrubber type being considered. Oxidation is defined as the percentage of net sulfur dioxide absorbed that is converted from sulfite or bisulfite to sulfate. Oxidation is an important factor in the economics of any scrubber type being considered, since increased oxidation results in increased electrical energy requirements. Consequently, a scrubber with excessive oxidation, by virtue of oxidation alone, could be economically unattractive.

To predict or estimate oxidation percentages, the relationship between scrubber design parameters and oxidation must be known. The best but most expensive way to obtain this information would be to run tests on various scrubber types. However, due to time and expense, testing was not possible for this study; thus, our inquiries to scrubber manufacturers requested that they supply us with any data they had on oxidation. We also made an extensive literature search for such information. None of the vendors was able to give us any specific information on oxidation, although one, The Norton Company, stated that oxidation would not exceed 2 percent.

Literature Search

Our literature search had two objectives. The first objective was to search for information concerning the mechanism of absorption of oxygen into sodium sulfite solutions and those factors affecting that absorption. The second objective was to relate the absorption mechanism to the chemical reaction and to see how factors which affect the reaction also affect the absorption into the liquid. Once this information had been collected, the task of relating it to absorber types could be done.

Absorption Mechanism and Chemical Reaction. The literature appears to be in agreement on the mechanisms of absorptions of oxygen into sodium sulfite solutions. It has been proved that the gas film resistance in this transfer is negligible.[4,6] Thus, the oxygen transfer mechanism can be separated into two individual rate processes:

1. Transfer of gas from gas-liquid interface to the bulk of the liquid
2. Chemical reaction of O_2 with sulfite

However, it has also been shown by Wartman[18] that the oxidation of solutions of normal sulfites is limited by the rate at which solutions will absorb oxygen. In Wartman's tests, all carried out at 25 C, 25 cc of distilled water were saturated with pure oxygen and then 2 cc of 3 molar ammonium sulfite were added, mixed, and allowed to stand for a given period of time. HCl was then added to stop further oxidation and the solution was analyzed for sulfates. The test was repeated for different time intervals varying from 7 sec to 3 min. Bisulfite solutions were also tested in this manner. All dissolved oxygen was consumed by normal sulfite within 7 sec, whereas with bisulfites the reaction proceeded much more slowly. Wartman also measured the amount of oxidation occurring when pure oxygen was bubbled through a solution of normal sulfite. By increasing the contact surface between gas and liquid, Wartman found that the rate of oxidation could be greatly increased. It then follows from this work that the rate-limiting oxygen transfer mechanism becomes the transfer of gas from gas-liquid interface to the bulk of the liquid.

Qualitatively, there are certain variables which affect the diffusional transfer or rate of absorption of oxygen.[4,6,9] These are:

1. Area of contact
2. Time of contact
3. Turbulence
4. Concentration gradient (magnitude of driving force)
5. Temperature

Even if there were no chemical reaction in the overall mechanism, the following factors which would affect chemical reaction also would affect the rate of absorption:[2,3,5,6,8,10,14,17]

6. Solute concentration
7. pH
8. Light intensity
9. Catalysis

Our literature search resulted in information showing how the above variables affect direction and intensity in the oxidation of sodium sulfite to sodium sulfate.

Area of Contact. The literature[18] states that oxidation increases as the area of contact increases. The oxygen transfer mechanism is a function of the transfer of gas from the gas-liquid interface to the bulk of the liquid and can be expressed as:

$$r_d = k_L (A)(c_i - c_L) \qquad (1)$$

where

r_d = rate of oxygen absorption, grams O_2 absorbed/(hr) (liter)

k_L = liquid-film mass transfer coefficient, grams O_2 transferred/(hr) (sq cm) (unit concentration difference)

A = transfer area, (sq cm)

$(c_i - c_L)$ = driving force of gas across interface into bulk of liquid, (grams/liter)

We have estimated that 5 percent oxidation would be expected from a Johnstone type scrubber, and the contract states that this figure should be used as a base case. On this basis, the effect of gas-liquid contact area on other types of scrubbers can be expressed as follows:

$$\text{Percent oxidation} = (\text{Base case oxidation})\left(\frac{\text{Contact area of new scrubber}}{\text{Contact area of base case scrubber}}\right) \qquad (2)$$

It must be kept in mind, however, that this estimate will be accurate only to the extent that the base case figure and the determination of the area of contact are exact.

Time of Contact. The literature made no definitive statements on time of contact. However, it did imply that as time of contact increased, oxidation increased. Unfortunately, there was no basis in the literature to show a quantitative relationship between the two.

Turbulence (Agitation). Since absorption is controlled by liquid film resistance, it can be inferred[4,6,9,15] that agitation of the liquid phase enhances the rate of mass transfer and increases oxidation. However, as in the case of time of contact, there was no basis in the literature for the development of a quantitative relationship.

Concentration Gradient (Magnitude of Driving Force). Referring to Equation 1:

$r_d = k_L (A)(c_i - c_L)$

c_i = concentration of oxygen at the interface, (grams/liter)

c_L = concentration of oxygen in the bulk of the liquid, (grams/liter)

The literature[4,9,15] defines the term $(c_i - c_L)$ as the driving force for absorption. Since oxygen solubility follows Henry's Law very closely and since the gas film coefficient is negligible, the partial pressure of oxygen at the interface is the same as that in the bulk of the gas.

$$c_i = P_{O_2}/H_{O_2}$$

P_{O_2} = partial pressure of solute in gas phase
H_{O_2} = Henry's law constant

Also, since the chemical reaction is very rapid, $c_L \simeq 0$ and the driving force is approximately equal to the concentration of oxygen at the interface. It can be seen from the above that by decreasing the oxygen content in the flue gas, oxidation should decrease proportionately.

It should be emphasized that operators of boiler plants using any process affected by oxidation should try to maintain excess air at the minimum consistent with good boiler operation. The exact values of excess air will depend on the type of coal and the design of the boiler system.

Temperature. Although published data on the solubility of oxygen in various solvents indicate that, as temperature increases, solubility of oxygen decreases, Cooper et al[6] state that the oxidation rate of sodium sulfite solutions increases by a factor of 2 when the temperature is increased from 32 F to 50 F. He also found that in the range of 68-104 F, the rate increases more slowly with increasing temperature. Our pilot plant data indicate that as temperature was increased in the range of 100-132 F, oxidation increased, although we found very little change in oxidation with temperature from 132 F to 140 F. A quantitative explanation for these findings is not available.

Solute Concentration The literature is not very clear on this subject. Several sources[10,18] contend that oxidation rate decreases as solute concentration increases. Other sources[6,8,19] indicate that oxidation proceeds at a rate independent of sulfite-ion concentration over concentration ranges as wide as 0.35-1.0 normal.

pH. One literature source[10] stated that oxidation apparently increased as acidity of the solution increased. This position is contested by several sources[5,6,8,13] who contend that oxidation is greatest in a neutral or slightly alkaline solution. We are inclined to accept the views of the latter since Wartman[18] noted that the oxidation of bisulfite, which has a pH of about 3.5, was slower than for neutral sulfite.

Light Intensity. It has been established that oxidation of sodium sulfite solutions to sodium sulfate is a chain reaction which is highly light sensitive.[2,8,14] However, the literature fails to provide an insight as to relative increases in oxidation rates due to the extreme light sensitivity of the reaction. This is not a factor in our process since our equipment is all metal. However, since light is a factor, laboratory tests done in glass might have shown higher oxidation than should be expected in a commercial plant.

Catalysis. It has also been shown that oxidation of sodium sulfite solutions is highly susceptible to both positive and negative catalysis. The heavy, variable-valence metal ions, such as Fe, Cu, Co, Ni and Mn have a powerful positive catalytic effect.[2,3,5,8,14,17] There are inhibitors which tend to reduce activity of these metal catalysts. Organic alcohols, phenols and potassium cyanide have a very strong negative effect.[2,8,10,14] However, if trace amounts of NO_x are present in the gas, oxidation becomes very difficult to inhibit.[14] A mechanism[1] has been proposed to explain both positive and negative catalysis, but has not been proved. The literature on catalysis as it affects oxidation yields little information which is quantitatively applicable.

Application of Literature Information

Of the nine identifiable factors affecting oxidation of sodium sulfite solutions to sodium sulfate, six are independent of scrubber design parameters. Temperature, solute concentration, pH, and light intensity will be common to all scrubber types. Concentration gradient and catalysis (positive) will be a function of boiler operations. Negative catalysis, if applicable, could be used in any scrubber, regardless of its type. This leaves only the factors of area of contact, time of contact, and turbulence as variables in the various scrubber designs.

Time of contact and turbulence, while unique to each scrubber type, were not determinable from the information provided. Therefore, only area of contact has been used in this report to estimate oxidation rates for various scrubber types.

As shown in Equation 2, these estimates will be only as accurate as the base case oxidation percentage and the determinations of area of contact. Even in pilot plant testing, Johnstone experienced wide ranging oxidation percentages. Because of this, oxidation rates shown in this report may be very inaccurate. It would seem, then, that testing programs are

justified to generate good design information on various scrubber types. This cannot be done within the scope of this report but could be a prerequisite to the final selection of a scrubber design for application in a commercial plant.

Estimated Oxidation for Scrubbers Studied

The results of pilot plant work and studies of Johnstone's work suggest that oxidation of sulfite to sulfate over Johnstone wood slat packing will be approximately 5 percent. We have attempted to make predictions of the percentage of sulfite oxidized in each of the several types of scrubbers studied. The figures are shown in Table 4. Using the wood slat packing as a base, with 5 percent oxidation expected, we estimate the other types of scrubber would produce the following results:

Type of Scrubber	*Percent Oxidation*
Wood slat packing	5
Intalox Saddle packed tower	10-15
Stone & Webster Ripple Tray	8-10
Glitsch grid impingement packing	3-5
Cyclonic contactor	2-5
Floating ball contactor	2-5
Venturi contactor	10-15

All except the venturi scrubber are based on a contact area relative to wood slat packing. Since no suitable method was found to determine the surface area of liquid droplets for the venturi contactor, we have used a value reported for a venturi unit not quoted (private communication).

The Norton Company had stated that oxidation would not exceed 2 percent, whereas calculations based on surface area indicate 15 percent. Since Norton's value is based on different operating conditions, it would require testing under our conditions to determine what the true value would be.

MATERIALS OF CONSTRUCTION

The sulfur dioxide recovery pilot plant in Tampa was constructed of materials that were available and thought to be suitable from a corrosion standpoint. The quench tower, the quench water pump casing, impeller shafts and wearing rings were 316 stainless steel. The scrubber recirculation pumps were also stainless steel. The absorption section was resin-impregnated fiberglass. Piping was stainless steel or polyvinyl chloride. The blower was resin-coated carbon steel. After several months of operation, the fiberglass venturi scrubber was replaced, because of high oxidation rates, with a 3 stage plexiglass tower using plastic spray nozzles for atomization of the recirculated sulfite solutions. The scrubber recirculation pumps were replaced with polyethylene pumps at the same time.

Operation of the pilot plant over a four month period showed that stainless steel was attacked slightly in the quench system. Thus, a better choice would be a nonmetallic material or rubber lined steel. As long as there is no carryover from the quench zone, mild corrosion would not affect the operability of the system. However, carryover would add metallic products of corrosion and dissolved metal ions from the fly ash to the absorption circuit and would eventually form metal hydroxides which must be removed from the feed stream before it enters the electrolytic cells. Because of problems associated with this possibility, lined equipment is a virtual necessity. The absorption section and internals should also be constructed of or lined with nonmetallic material. This arrangement would favor designs using nonmetallic packings assuming that nonmetallic or coated supports and distributors could be fabricated. Designs with large exposed metal surfaces, such as multivane cyclonic contactors, would be more difficult to coat and would experience more erosion from particulate matter and subsequently more corrosion.

We have relocated the blower to the inlet side of the quench tower where hot gas will be handled and special materials for the blower are not required. This will require a larger blower, but will eliminate the structure previously required to support the blower and will make maintenance easier and less expensive.

Piping should also be constructed of or lined with nonmetallic material. Polyvinyl chloride piping was satisfactory for the pilot plant. Epoxy resin fiberglass should be suitable for very large piping.

Duct work for the hot gases would be of carbon steel but should be lined or coated with nonmetallic coating after the quench section.

ECONOMIC COMPARISON OF VARIOUS SCRUBBERS FOR THE STONE & WEBSTER/IONICS PROCESS

INVESTMENT

The cost of a scrubber system based on the work of Johnstone was estimated, to serve as the base cost of the scrubber system in the S&W/Ionics sulfur dioxide removal and recovery process. The estimate was made for a four boiler 1,200 Mw station. It included:

1. Duct work required to take hot flue gas from existing precipitators to quench devices, which would also act as solids scrubbers
2. Four gas quenchers
3. Four two stage sulfur dioxide scrubbers
4. Fans to provide a pressure boost equal to the overall system pressure loss
5. Duct work to carry the flue gas from the scrubbers back to the stack
6. All necessary pumps for quencher and absorption sections to provide liquid flows as required for each unit
7. Fired reheaters in the quenched, sulfur dioxide cleaned gas to provide enough reheat to prevent condensation in the stack
8. Construction costs and Contractor's field costs, engineering burdens, and fees

Investment costs were estimated for each of the other six types of scrubbers for which manufacturers supplied information. Each estimate considered the same eight cost categories shown above. In addition to these items, the effect of oxidation on the number of cells required to process sodium sulfate was estimated and, depending on whether the new design was more or less prone to oxidation as compared to the Johnstone type, a capital adjustment was made to the estimate for each type of scrubber.

Where oxidation is shown as a range, the lowest estimated level was used to compute the investment penalty or credit.

The estimates for the seven types of scrubbers are shown in Table 5. All six of the scrubber systems have been estimated to cost more than the Johnstone based system

originally used by S&W in its proposal to NAPCA. The Johnstone, Glitsch grid, Intalox Saddle and S&W Ripple Tray systems can be scaled up to the point where only one dual purpose, quench, two stage scrubber tower need be built for each boiler. Since only one tower is required for each boiler, the cost of duct work for these systems is significantly smaller than the cost of duct work in the more complex venturi, cyclonic, and floating ball systems.

SCRUBBER SYSTEM OPERATING COST

One of the major operating costs in the scrubber systems is the cost of power for the flue gas fans which are used to overcome the sulfur dioxide scrubber system pressure drops. The venturi system, while it has a lower ΔP in the gas stream than the Johnstone system, uses the quench and absorption fluids as drivers to boost pressure through the scrubbing device. In spite of this interesting operational characteristic, the total energy required for fans and fluid pumps in the venturi system is slightly higher than that required by the Johnstone unit. All of the other systems studied also have a higher energy cost than the Johnstone system.

While the oxidation levels in all of the scrubber types studied can only be estimated at this time, it would appear that the cyclonic (Pulverizing Machinery) and floating ball (Buell) could well generate lower oxidation levels than the Johnstone type. The Glitsch system might also offer slight reductions as compared to the Johnstone unit. In addition to the operating cost associated with pressure drop in the scrubber system, a sum would be needed to cover the operating cost of an additional (or reduced) number of cells required to cope with estimated increases (or decreases) in the oxidation level characteristic of each type.

Maintenance costs were assumed to be 2 percent of the installed cost.

A summary of the operating costs is shown in Table 6.

SULFUR DIOXIDE SCRUBBER SYSTEM COST EVALUATION

The basis of evaluation for this study was the net present value of the cash flows for each type of scrubber over a 15 yr period, using a discount factor of 6 percent. The capital return factor used was 17.50 percent. This includes bonded debt cost, depreciation on a 5 percent straight-line basis, local taxes and insurance, and utility return on equity after taxes (see Appendix B).

Since the construction period for recovery plants will be short, we have assumed that the total differential investment will be made in the first year of operation. This assumption will tend to decrease the effect of capital cost and overemphasize the value of operating cost savings. The formula used, 10.288 (0.1750 Δ Investment + Δ Operating) equals the 15 yr evaluated cost of the scrubber. The Johnstone unit 15 yr evaluated cost is zero. All of the other units are considerably more expensive than the Johnstone unit. Based on the estimates made, it would appear that only the Glitsch grid packed tower would offer any competition to the Johnstone system proposed by S&W for inclusion in an add-on sulfur dioxide recovery situation suggested in our proposal to NAPCA.

The evaluated comparative costs of the six types of scrubbers studied are shown in Table 7.

Until the manufacturers of venturi, cyclonic, and floating ball units can produce very large single units, the evaluated cost of such units makes their use in the S&W/Ionics process impractical.

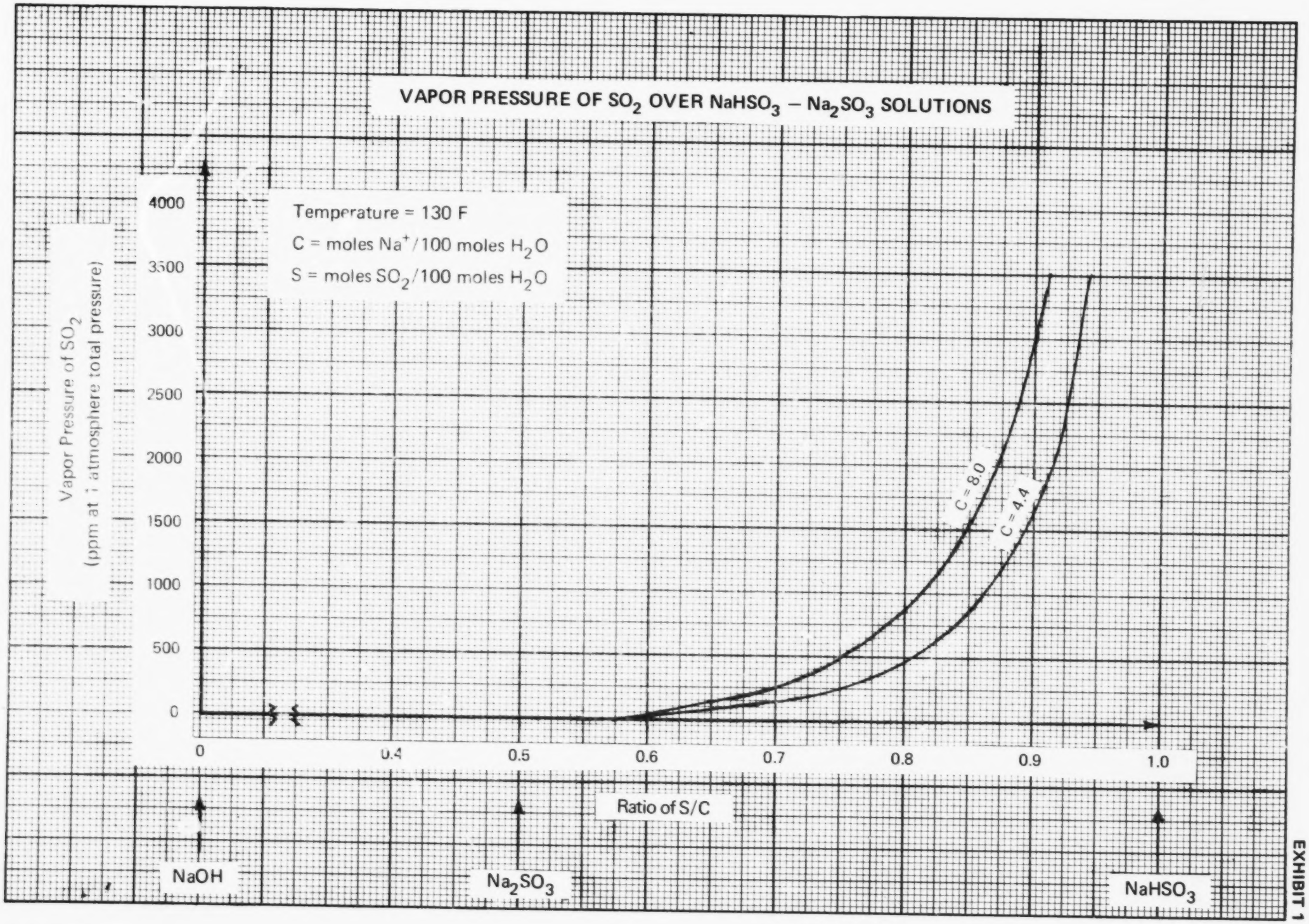
VAPOR PRESSURE OF SO_2 OVER $NaHSO_3$ – Na_2SO_3 SOLUTIONS
Temperature = 130 F
C = moles Na^+/100 moles H_2O
S = moles SO_2/100 moles H_2O
Vapor Pressure of SO_2
(ppm at 1 atmosphere total pressure)
4000
3500
3000
2500
2000
1500
1000
500
0
C = 8.0
C = 4.4
0
0.4
0.5
0.6
0.7
0.8
0.9
1.0
Ratio of S/C
NaOH
Na_2SO_3
$NaHSO_3$

EXHIBIT 1

CALCULATED VALUES OF P_{SO_2} USED FOR EXHIBIT 1

$$P_{SO_2}\ (ppm) = \left[\frac{MC}{7.6 \times 10^{-4}}\right]\left[\frac{(2\ S/C - 1)^2}{(1 - S/C)}\right]$$

At 130 F, M = 0.0452

S/C	*(At C = 4.4) P_{SO_2} (ppm)*	*(At C = 8.0) P_{SO_2} (ppm)*
0.60	26	48
0.65	67	122
0.70	140	254
0.75	262	476
0.78	373	678
0.80	471	856
0.82	595	1083
0.84	756	1375
0.86	969	1762
0.88	1260	2290
0.89	1447	2632
0.90	1675	3045
0.91	1955	3555
0.92	2308	–
0.93	2765	–
0.94	3377	–

CALCULATED VALUES OF P_{SO_2} USED FOR EXHIBIT 1

$$P_{SO_2}\ (\text{mm Hg}) = \left[\frac{MC}{7.6 \times 10^{-4}}\right]\left[\frac{(2\,S/C - 1)^2}{(1 - S/C)}\right]$$

At 130 F, M = 0.0452

S/C	*(At C = 4.4) P_{SO_2} (mm Hg)*	*(At C = 8.0) P_{SO_2} (mm Hg)*
0.60	26	48
0.65	67	122
0.70	140	254
0.75	262	476
0.78	373	678
0.80	471	856
0.82	595	1083
0.84	756	1375
0.86	969	1762
0.88	1260	2290
0.89	1447	2632
0.90	1675	3045
0.91	1955	3555
0.92	2308	
0.93	2765	
0.94	3377	

APPENDIX B

DERIVATION OF ECONOMIC FACTORS

1. **Annual Capital Return Factors**

Bonded Debt, 60% I at 6%	(3.6%) (I)
Depreciation, 20 yr Straight Line	(5.0%) (I)
Local Taxes and Insurance	(3.0%) (I)
Utility Return on Investment	
40% equity at 7% allowable after tax	(2.8%) (I)
Federal Tax (52.2% rate)	(3.1%) (I)
Total Return on Annual Basis	(17.50%)(I)

2. **Annual Cost Differential of Any Process – Johnstone as Base**

Annual cost of operation will have two components:

A. Return factor which will be 17.50 percent of differential investment between Johnstone and other types

B. Operating Cost Component which will be differential between Johnstone and other types

If it is assumed that the load factor on the recovery plant in later years, i.e., 10 through 15, is equal to the early years, i.e., 1 through 5, the present worth of any alternate is equal to the discounted sum of ($0.1750\ \Delta I + \Delta$ Operating) over a 15 yr period when the expression is calculated for the first year. The discount factors when summed equal 10.288.

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Table 1

FLY ASH REMOVAL
LIST OF VENDORS CONTACTED
STUDY OF SO_2 SCRUBBERS
NATIONAL AIR POLLUTION CONTROL ADMINISTRATION

	Quoted	*Reason for Declining to Quote*
Pulverizing Machinery Company (Formerly Aireton Eng. Co.)	8-22-69	
Buell Engineering Company, Inc.	9-11-69	
Buffalo Forge Company		Equipment too small
Burgess-Manning Company		Do not have suitable equipment
The Ceilcote Company, Inc.	9-19-69	
Chemical Construction Corporation		Will not quote without fee
Clermont Engineering Co.		Do not have equipment this size
Corning Glass Works		Equipment too small
Croll Reynolds Company, Inc.	8-27-69	
Dorr-Oliver		Declined to quote
The Ducon Company, Inc.	9-18-69	
Fuller Company		Do not offer this type of equipment
Fritz W. Glitsch & Sons, Inc.	9-17-69	
Heil Process Equipment Corporation	8-25-69	
Maurice A. Knight Co.	7-18-69	
Koch Engineering Company, Inc.	9-10-69	
National Dust Collector Corporation		Can quote only in connection with Combustion Engineering
Nooter Corporation		Can offer no information on flue gas treating
Peabody Engineering Corporation	9-16-69	
Research Cottrell, Inc.	7-19-69	Quote on fly ash removal only
U.O.P. Air Correction Division		See Note below
Claude B. Schneible Co.	8-11-69	
Norton Company (Formerly U. S. Stoneware, Inc.)	9-15-69	

Note: The UOP Air Correction Division declined to quote because they believe that low liquid velocities and very high retention times are required in producing a liquor high in bisulfite. This is in opposition to the operating principle of their Turbulent Contact Absorber (TCA) scrubber.

VENDORS QUOTING ON FLY ASH REMOVAL AND QUENCH EQUIPMENT
STUDY OF SO_2 SCRUBBERS
NATIONAL AIR POLLUTION CONTROL ADMINISTRATION

Manufacturer	*Type of Fly Ash Removal Section*					
	Tray	*Venturi*	*Cyclone*	*Impinge-ment*	*Packed Bed*	*Other*
Pulverizing Machinery Company (Formerly Aireton Eng. Co.)		X				
Buell Engineering Company, Inc.						X
The Ceilcote Company, Inc.					X	
Croll-Reynolds Co., Inc.		X				
Fritz W. Glitsch & Sons, Inc.				X		
Heil Process Equipment Corporation		X				
Maurice A. Knight Co.		X				
Koch Engineering Company, Inc.		X				
Peabody Engineering Corporation				X		
Research-Cottrell, Inc.						X
Claude B. Schneible Co.			X			
Norton Company (Formerly U. S. Stoneware, Inc.)					X	

Table 2

VENDORS QUOTING ON SO_2 SCRUBBERS
STUDY OF SO_2 SCRUBBERS
NATIONAL AIR POLLUTION CONTROL ADMINISTRATION

Manufacturer	*Type of SO_2 Absorption Section*					
	Tray	*Venturi*	*Cyclone*	*Impinge-ment*	*Packed Bed*	*Other*
Pulverizing Machinery Company (Formerly Aireton Eng. Co.)			X			
Buell Engineering Company, Inc.						X
The Ceilcote Company, Inc.					X	
Croll-Reynolds Company, Inc.		X				
Fritz W. Glitsch & Sons, Inc.	X					
Heil Process Equipment Corporation		X				
Maurice A. Knight Co.					X	
Koch Engineering Company, Inc.					X	
Peabody Engineering Corporation				X		
Claude B. Schneible Co.			X			
Norton Company (Formerly U. S. Stoneware, Inc.)					X	

Table 3

SUMMARY OF VEND
STUDY OF SO_2
NATIONAL AIR POLLUTION C

Vendor	Pulverizing Machinery Co.	Buell Eng. Co.	The Ceilcote Co., Inc.	Croll-Reynolds Co., Inc.	The Ducon Co., Inc.	Fritz W. Glitsch & Sons, Inc.	Heil Process Equipment Corp.
Quench Section							
Type of Quench	Venturi	Floating Ball	Tellerette Packing	Venturi	Venturi	Spray	Water-Jet
Number of Units in Parallel	7	9	8	3	12	1	8
Size of Unit	6'-6" diam x 11'	14'-0" diam x 27'	20' x 14' x 18'	10' diam x 40'	9'-2" diam x 13'-6"	38' diam x 20'	Not stated
Materials of Construction	Steel-Coated Emersite	Fiberglass	Reinforced Plastic	Steel with Amercoat	316 S.S.	Steel – Rubber Lined	316 S.S.
Fly Ash Removed, %	37 > 1	95	98-99 > 5μ +	Most	99+	Not stated	Most - Over 5μ
Pressure Drop, In. H_2O	10	4	1	0	10	--	3
Water Circ. Rate, Gpm (Total)	725	1,760	6,800	7,500	9,600	1,600	Not stated
SO_2 Scrubbing Section							
Type of Scrubber	Cyclonic	Floating Ball	Tellerette Packing	Venturi	Cyclonic Multivane	Impingement Grids-410 S.S.	Packed Section
Number of Units in Parallel	7	9	8	3	12	1	8
Total Number of Units	14	9	8	6	12	1	8
Size of Unit	12'-6" diam x 44'	14' diam x 35'	Included in Quench Sec.	10' diam x 40'	12' diam x 28'-3"	38' diam x 34'	Not stated
Materials of Construction	Steel – Rubber Lined	316 S.S.	Reinforced Plastic	Fiberglass Dynel Liner	Steel-Epoxy Lined	Steel – Amercoat	Plastic
Total Packed Height, Ft	–		–	–	–	16	Not stated
Stages SO_2 Removal	2	2	2	2	1	2	Not stated
SO_2 Removed, %	95	Not stated	95-97	Not stated	90-92	Not stated	99
Pressure Drop, In. H_2O	12	7.5	5	0	3	7.0	Not stated
Liq. Recirc. Rate, Gpm (Total)	14,700	3,520	4,000	10,000	6,000	3,400	Not stated
Estimated Oxidation, %	2-5	2-5	(2)	10-15	(2)	3-5	(2)

Notes: (1) Quotations based on equipment required for one boiler for 300 megawatt unit; (2) Not Est

Table 4

Y OF VENDORS' QUOTATIONS[1]
DY OF SO_2 SCRUBBERS
LLUTION CONTROL ADMINISTRATION

il Process quipment Corp.	Maurice A. Knight Co.	Koch Engineering Co., Inc.	Peabody Eng. Corp.	Research-Cottrell, Inc.	Claude B. Schneible	Norton Co.	S&W Eng. Corp.	S&W Eng. Corp.
ater-Jet	Venturi	Venturi	Impingement Baffle Plate	Flooded Disc	Multiwash Cyclonic	Packed 3" Intalox Saddles	Spray	Spray
	6	4	4	4	14	1	1	1
ot stated	Not stated	12' diam x 20'	12' x 30' x 25'	8'diam x 13'-9"& 21'diam x 33'	10'-9" diam x 26'	43' diam x 20'	37'-6" diam x 20'	37' diam x 20'
6 S.S.	Steel Lined	Steel – Rubber Lined	316 S.S.	Steel – PVC Lined	304 S.S.	Steel – Amercoat	Steel – Amercoat	Steel – Amercoat
ost – Over 5μ	95	90	Not stated	98	99 > 1	99+	90+	90+
	20	30	–	6	–	0.5	–	–
ot stated	6,600	6,000	3,040	4,400	3,400	4,350	1,600	1,600

ecked ection	Packed 2" Saddles	Packed 3½" P.P. Rings	Impingement Baffle Plate	No Quote	Multiwash Cyclonic	Packed 3" Intalox Saddles	Ripple Trays	Johnstone Slat Packing
	1	2	4	No Quote	14	1	1	1
	1	2	4	No Quote	14	1	1	1
ot stated	Not stated	30' diam x 56'	Included Above	No Quote	10'-9" diam x 22'	43' diam x 42'	37'-6" diam x 40'	37' diam x 28'
astic	Steel – Plastic Lined	316 S.S. & Steel – Rubber Lined	316 S.S.	No Quote	304 S.S.	Steel – Amercoat	Steel – Amercoat	Steel – Amercoat
ot stated	Not stated	14	–	No Quote	–	24	(10 trays)	10
ot stated	Not stated	1	3	No Quote	1	2	2	2
g	Not stated	Not stated	Not stated	No Quote	95	Not stated	Est. 90	Est. 90
ot stated	4	8	11 (Total)	No Quote	6.8 (Total)	8.5	18	4 (Total)
ot stated	3.5 Gpm/S.F.	5,000	4,080	No Quote	3,400	8,700	4,600	3,400
2)	(2)	(2)	(2)	No Quote	(2)	10-15	8-10	5

tt unit; (2) Not Estimated

B

Table 5

INVESTMENT COMPARISON
SO_2 SCRUBBERS FOR THE STONE & WEBSTER/IONICS PROCESS
NATIONAL AIR POLLUTION CONTROL ADMINISTRATION

Type of Absorber	*System Cost,[1] 1,200 Mw*	*Differential from Johnstone Base Cost*
Johnstone wood slat packing	$6,040,000	0
Glitsch grid impingement packing	6,720,000	+$680,000
Intalox Saddle packing	7,920,000	+1,880,000
Venturi scrubbers	8,960,000	+2,920,000
Stone & Webster Ripple Tray	9,500,000	+3,460,000
Cyclonic scrubbers	11,920,000	+5,880,000
Floating ball	12,800,000	+6,760,000

[1] *Cost includes quench, sulfur dioxide scrubber, pumps for two stages, blowers, ducts, piping, instrumentation, dampers, civil accounts, investment in cells for oxidation differential from Johnstone base for four 300 Mw boilers.*

OPERATING COST COMPARISON
SO_2 SCRUBBERS FOR THE STONE & WEBSTER/IONICS PROCESS
NATIONAL AIR POLLUTION CONTROL ADMINISTRATION

Type of Scrubber	*Scrubber System Differential Power Cost, $/Year*	*Oxidation Differential Power Cost, $/Year*	*Maintenance Differential $/Year (2%ΔI)*	*Total Operating Cost Differential, $/Year*
Johnstone	–	–	–	–
Glitsch grid	+$18,800	-$84,200	+$13,600	-$41,800
Intalox Saddle	+42,800	+100,500	+37,600	+190,900
Venturi	+20,400	+110,500	+58,400	+189,300
Ripple Tray	+138,800	+126,300	+69,200	+334,300
Cyclonic	+212,000	-126,300	+117,600	+203,300
Floating ball	+109,000	-126,300	+135,200	+117,900

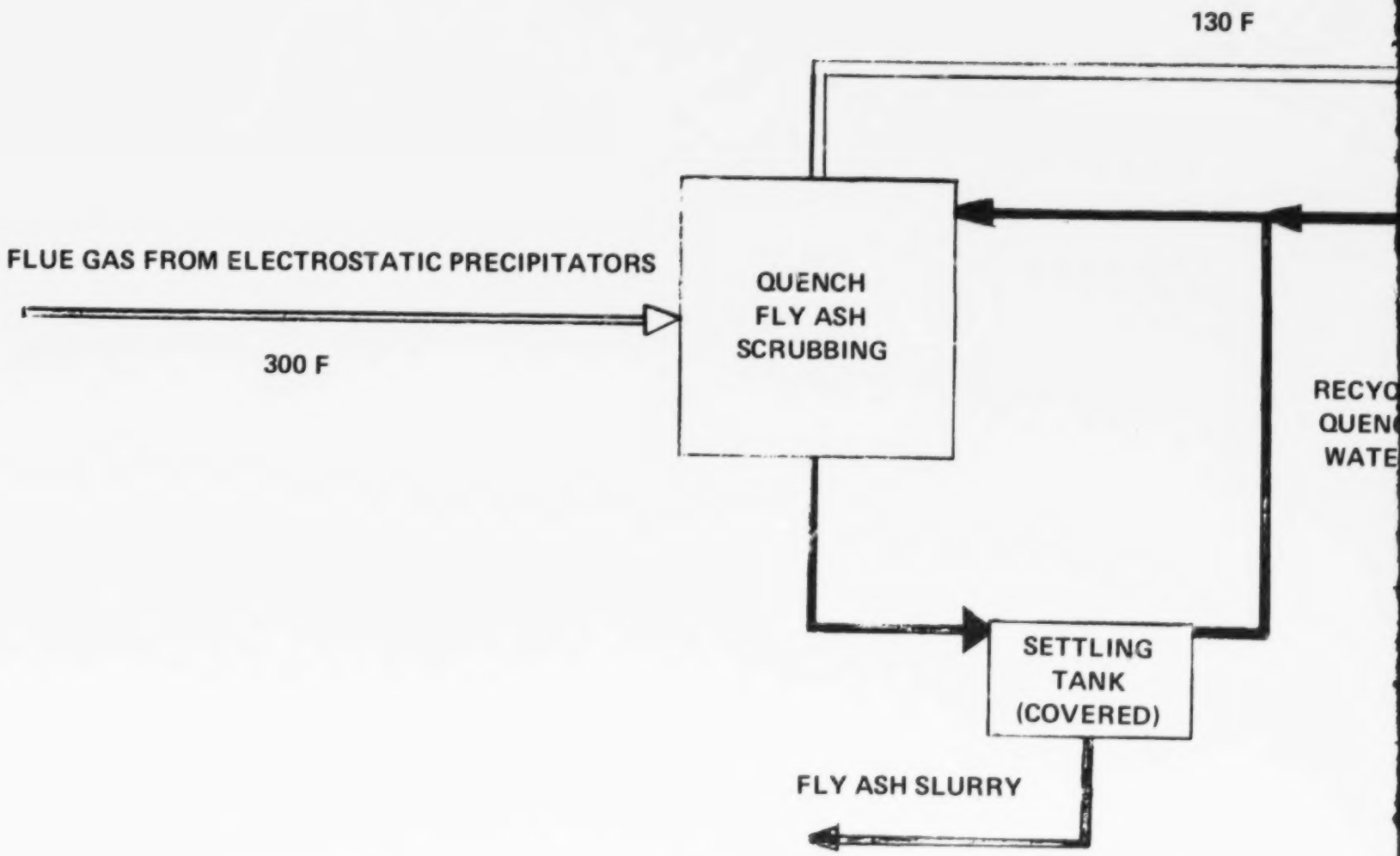
Note: The annual operating cost of a Johnstone-based system will be on the order of $2,500,000.

Table 6

Table 7

COST EVALUATION
SO_2 SCRUBBER SYSTEM
NATIONAL AIR POLLUTION CONTROL ADMINISTRATION

Type of Scrubber	*Comparative Present Worth, Johnstone Type as Base*
Johnstone wood slat packing	Base (Zero)
Glitsch grid impingement packing	+$794,000
Intalox Saddle packing	+5,350,000
Venturi scrubbers	+7,200,000
Stone & Webster Ripple Tray	+9,670,000
Cyclonic scrubbers	+12,680,000
Floating ball	+13,380,000


130 F
FLUE GAS FROM ELECTROSTATIC PRECIPITATORS
300 F
QUENCH
FLY ASH
SCRUBBING
RECYC
QUEN
WATE
SETTLING
TANK
(COVERED)
FLY ASH SLURRY

A

30 F

QUENCH
WATER
MAKEUP

RECYCLE
QUENCH
WATER

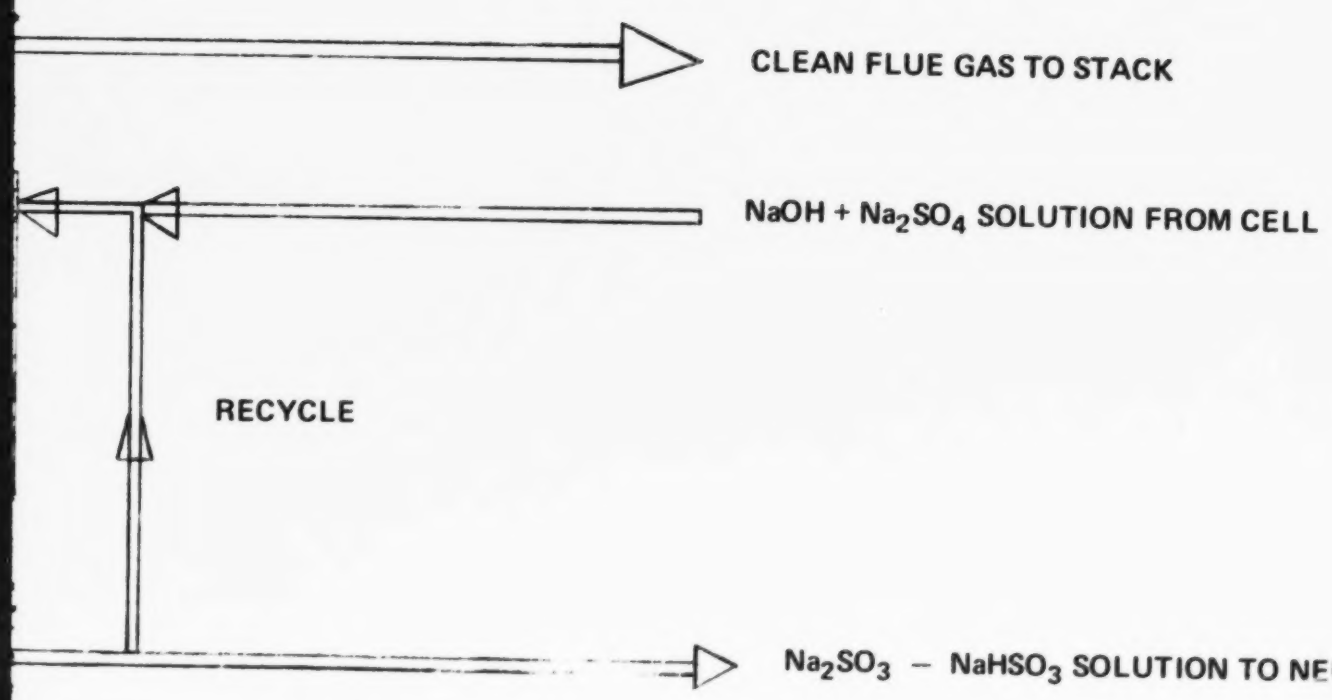
SO_2
SCRUBBING

RECYCLE

B

7 6 5 4 3 2 1

FIGURE 1



C

SCHEMATIC FLOW DIAGRAM
SO_2 REMOVAL
STUDY OF SO_2 SCRUBBERS
NAPCA

STONE & WEBSTER ENGINEERING CORPORATION